

1 **PRODUCTION OF HIGH QUALITY**
2 **LUBRICANT BRIGHT STOCK**

3
4 FIELD OF THE INVENTION

5
6 The present invention relates to a process for producing a high quality
7 lubricant bright stock from heavy petroleum feedstocks.

8
9 BACKGROUND OF THE INVENTION

10
11 This invention is directed to a process for preparing a high quality lubricant
12 bright stock from heavy petroleum feedstocks. These heavy feedstocks are
13 often contaminated with sulfur, nitrogen, asphaltenic and metal contaminants,
14 which must be removed in preparing the lubricant base stock. They also
15 generally contain significant amounts of waxy materials.

16
17 Low valued oils such as deasphalted oil (DAO) are increasingly being
18 hydrotreated and used as FCC cracker feed to produce gasoline. Severity of
19 deasphalting is much less for making fuels than for making lubricant bright
20 stock. Consequently, the purity of fuels-application DAO is too low to make
21 lubricant bright stock with adequate stability for use in finished lubricant
22 applications. However, increasingly stringent mandated limits on gasoline
23 sulfur require higher severity in fuels DAO hydrotreating. These changes
24 improve the quality of DAO as a feed to hydrocracking to produce high
25 viscosity lubricant base oils. Nevertheless, the highest boiling portion of DAO
26 contains high molecular weight waxes which are difficult to remove, leading to
27 low yields of high cloud point products. The highest boiling fractions also
28 contain large polycyclic molecules, which are difficult to completely saturate in
29 hydrofinishing and which lead to stability problems.

30
31 Conventional high quality Group II lubricant neutral oils having excellent
32 oxidation stability, good low temperature properties and high viscosity indices
33 are generally made by hydrocracking gas oils, followed by dewaxing and

1 optionally mild hydrotreating. Lubricant base stocks having viscosities of up to
2 about 100 cSt, measured at 40°C, are made in this manner. Higher viscosity
3 oils, for example, bright stocks and similar oils with viscosities of 220 cSt or
4 greater, are generally not made by isomerization dewaxing and generally do
5 not possess the high quality of Group II base oils prepared by isomerization
6 dewaxing. High viscosity oils of improved quality are in general demand,
7 especially for non-engine oil applications, such as industrial oils.

8
9 Such high viscosity oils generally require some bright stock in their
10 formulation, the amount of which depends on the product. In typical
11 formulations, Group I bright stock is used, which degrades the product when
12 blended with neutral oil. One problem with the quality of bright stock is that it
13 is not a distillate and is typically of low quality, particularly with respect to
14 oxidation stability. Thus, there is a need for a method for producing an
15 oxidation stable, good quality lubricant bright stock.

16
17 In addition, feedstocks which are useful for making lubricant bright stock have
18 generally been limited to gas oils, and specifically vacuum gas oils. Residuum
19 streams are generally difficult to process for lubricant base oils. Not only are
20 the sulfur, nitrogen and aromatic contents of residuum streams very high, but
21 the waxy materials present in these residuum streams are difficult to process
22 in the production of low pour point base oil products. It is especially, therefore,
23 to be able to produce good quality lubricant bright stock from
24 residuum-derived streams.

25 26 SUMMARY OF THE INVENTION

27
28 The present invention relates to a process for producing a lubricant bright
29 stock from a very heavy feed, having a normal boiling end point within the
30 range of 1150°F to 1300°F. The very heavy feed is obtained from petroleum
31 crude. The bright stock produced by the present process has a reduced cloud
32 point and better oxidation stability relative to bright stocks prepared by
33 conventional methods.

1 The present invention provides a process for producing a stable lubricant
2 bright stock comprising the steps of providing a petroleum residuum-derived
3 stream having a sulfur content of less than 1% and a nitrogen content of less
4 than 0.5%; separating the residuum-derived stream at a distillation cut point in
5 the range of 1150°F to 1300°F, into a heavy fraction and at least one light
6 fraction; hydrocracking the at least one light fraction under lube hydrocracking
7 in a lube hydrocracking zone in the presence of a hydrocracking catalyst and
8 hydrogen under conditions to reduce the concentration of sulfur and nitrogen
9 to suitable levels for hydroisomerization dewaxing; and dewaxing at least a
10 portion of the hydrocracked stream in a hydroisomerization zone in the
11 presence of an isomerization catalyst and hydrogen under hydroisomerization
12 conditions to produce a lubricant bright stock.

13

14 In one embodiment, a vacuum residuum fraction, which is optionally
15 hydrotreated/hydrocracked prior to further treating, is separated in a high
16 temperature fractionation step, at a cut point temperature in the range of
17 1150°F to 1300°F, into at least a heavy fraction and a light fraction. The light
18 fraction is further processed by hydroisomerization dewaxing to prepare a low
19 haze bright stock.

20

21 Conventional methods for preparing bright stock by hydroisomerization
22 dewaxing a vacuum residuum fraction, a hydrocracked vacuum residuum
23 fraction, deasphalted oil or hydrocracked deasphalted oil generally produces a
24 bright stock with unacceptable haze-forming tendencies.

25

26 Among other factors, the present invention is based on the surprising
27 discovery that the high temperature fractionation, with a cut point in the range
28 of 1150°F to 1300°F, concentrates the haze-forming components found in the
29 vacuum residuum fraction in the heavy fraction, and provides a low haze, and
30 relatively low sulfur containing light fraction which can be processed using
31 conventional methods for preparing a bright stock lubricant. The preferred
32 method of fractionating at the high temperatures required to achieve the
33 desired separation is a short path distillation, such as a wiped film evaporator.

DETAILED DESCRIPTION

Feed Stream:

The petroleum feed stream which is treated in the present invention is a residuum fraction, derived from the fractionation of a petroleum feed, preferably a crude feed. Other feeds which may be treated in the present invention include deasphalted oil, heavy coker products, and the like.

Deasphalted oil (DAO) may be recovered from a conventional deasphalting process, such as a solvent deasphalting process. Such processes are well known in the art. A process for preparing a DAO which is useful in the present invention is described, for example, in U.S. Patent No. 6,001,886, which is incorporated by reference. In a deasphalting process, a residuum is subjected to counter-current contacting at solvent deasphalting conditions, generally at a temperature in the range of 50°F to 400°F, preferably 150°F to 300°F, a dosage of from 0.5 to 10, preferably 1.0 vol. to 3.0 vol. solvent/vol. oil and a pressure of atmospheric pressure to 400 psig, preferably atmospheric pressure to 50 psig. The actual deasphalting conditions chosen are dependent on the solvent. That is, the temperature chosen should not exceed the critical temperature of the solvent and the pressure is maintained above the autogenous pressure to prevent vaporization. Deasphalted oil and solvent are removed by distillation or by stripping the asphalt layer, leaving behind a viscous asphaltic residue. Deasphalting solvents which are useful for this purpose include C₂ to C₈ paraffins, furfural and N-methyl-2-pyrrolidone. Propane and butane are preferred. Pentane is the most suitable solvent if high yields of deasphalted oil are desired. These lower-boiling paraffinic solvents may also be used as mixtures with alcohols such as methanol and isopropanol. Propane as a solvent results in the lowest yield of deasphalted oil and highest yield of asphaltic residue. Because propane is the preferred commercial solvent, the process is often referred to as propane deasphalting. Iso-butane and n-butane are also used commercially. The Rose (Residual Oil Solvent Extraction) process has been the object of many patents disclosing

different operating conditions, or the use of several solvents as specified, for example, in U.S. Patent Nos. 3,830,732 and 4,125,459. The preferred solvent in the Rose process is pentane. The Rose process includes a step under supercritical conditions adapted to separate the solvent from the deasphalted oil.

Hydrotreating/Hydrocracking the Residuum Feed:

The residuum feed stream is optionally upgraded as necessary prior to separation in the deep cut distillation stage. The feed to the deep cut distillation should have a sulfur content of less than 1% and a nitrogen content of less than 0.5%. The method of upgrading depends, at least in part, on the quality of the feedstock and the quality of the desired bright stock lubricant product. For example, the feed stream may desirably be hydrotreated to remove sulfur without extensive molecular weight conversion by hydrocracking. As used here, the upgrading step prior to separation is identified as a hydrotreating/hydrocracking step to indicate the range of upgrading severities which may be used in the present process.

Hydrotreating refers to a catalytic process, usually carried out in the presence of free hydrogen, in which the primary purpose is the desulfurization and/or denitrification of the feed stock. Generally, in hydrotreating operations, cracking of the hydrocarbon molecules (i.e., breaking the larger hydrocarbon molecules into smaller hydrocarbon molecules) is minimized and the unsaturated hydrocarbons are either fully or partially hydrogenated. Hydrotreating conditions include a reaction temperature between 400°F to 900°F (204°C to 482°C), preferably 650°F to 850°F (343°C to 454°C.); a pressure between 500 psig to 5000 psig (pounds per square inch gauge) (3.5 MPa to 34.6 MPa), preferably 1000 psig to 3000 psig (7.0 MPa to 20.8 MPa); a feed rate (LHSV) of 0.5 hr⁻¹ to 20 hr⁻¹ (v/v); and overall hydrogen consumption 300 to 2000 standard cubic feet per barrel of liquid hydrocarbon feed (53.4-356 m³ H₂/m³ feed). The hydrotreating catalyst for the beds will typically be a composite of a Group VI metal or compound

1 thereof, and a Group VIII metal or compound thereof supported on a porous
2 refractory base such as alumina and silica-alumina. Examples of
3 hydrotreating catalysts are alumina supported cobalt-molybdenum, nickel
4 sulfide, nickel-tungsten, cobalt-tungsten and nickel-molybdenum. Typically,
5 such hydrotreating catalysts are presulfided.
6
7 Hydrocracking is a process of breaking larger hydrocarbon molecules into
8 smaller ones. It can be affected by contacting the particular fraction or
9 combination of fractions, with hydrogen in the presence of a suitable
10 hydrocracking catalyst. The hydrocracking step reduces the size of the
11 hydrocarbon molecules, hydrogenates olefin bonds, hydrogenates aromatics,
12 opens rings, and removes traces of heteroatoms. Typical hydrocracking
13 conditions include: reaction temperature, 400°F to 950°F (204°C to 510°C),
14 preferably 650°F to 850°F (343°C to 454°C); reaction pressure 500 psig to
15 5000 psig (3.5 MPa to 34.5 MPa), preferably 1500 psig to 3500 psig
16 (10.4 MPa to 24.2 MPa); liquid hourly space velocity (LHSV),
17 0.1 hr⁻¹ to 15 hr⁻¹ (v/v), preferably 0.25 hr⁻¹ to 2.5 hr⁻¹; and hydrogen
18 consumption 500 to 2500 standard cubic feet per barrel of liquid hydrocarbon
19 feed (89.1-445 m³ H₂/m³ feed). The hydrocracking catalyst generally
20 comprises a cracking component, a hydrogenation component, and a binder.
21 Such catalysts are well known in the art. The cracking component may
22 include an amorphous silica-alumina phase and/or a zeolite, such as a Y-type
23 or USY zeolite. The binder is generally silica, alumina or silica-alumina. The
24 hydrogenation component will be a Group VI or Group VIII metal or oxides or
25 sulfides thereof, preferably one or more of molybdenum, tungsten, cobalt, or
26 nickel, or the sulfides or oxides thereof. If present in the catalyst, these
27 hydrogenation components generally make up from about 5% to about
28 40% by weight of the catalyst. Alternatively, platinum group metals, especially
29 platinum and/or palladium, may be present as the hydrogenation component,
30 either alone or in combination with the base metal hydrogenation components
31 molybdenum, tungsten, cobalt, or nickel. If present, the platinum group metals
32 will generally make up from about 0.1% to about 2% by weight of the catalyst.

1 **Deep Cut Distillation:**

2

3 Once a suitable residuum-derived petroleum feedstock has been obtained, it
4 is next separated into a heavy fraction and a light fraction by a deep cut
5 distillation. The deep cut distillation separates the heavy feed stream at a cut
6 point in the range of 1150°F to 1300°F into a heavy fraction and a light
7 fraction, the latter of which undergo further separation by distillation. The cut
8 point is the temperature at which there are equal amounts of material
9 overlapping from adjacent cuts. When data is not available for one or both
10 adjacent cuts, cut point estimates are the 10 and 90 percent points of the
11 distillation curve. The heavy fraction has a boiling point predominantly above
12 the cut point and the at least one light fraction has a boiling point
13 predominantly below the cut point. The heavy fraction may be used as feed to
14 the FCC or recycled to the hydrocracker. At least a portion of the light fraction
15 is used as a feedstock for hydrocracking discussed more fully below.

16

17 Special care is required to separate very high boiling materials in order to
18 minimize product degradation. WO 00/11113, which is incorporated herein by
19 reference, describes the use of special packing, stream stripping and high
20 vacuum to achieve high temperature separations without product degradation.
21 Other commercially available methods employ techniques developed as
22 molecular distillation methods. These are described in detail in, for example,
23 G. Burrows, *Molecular Distillation*, Oxford: Clarendon Press, 1960. Such
24 short-path distillation methods include falling film evaporators and wiped film
25 evaporators. An example of short path distillation is described in
26 U.S. Patent No. 4,925,558, incorporated herein by reference.

27

28 The American Society for Testing and Materials (ASTM) has established
29 guidelines for simulated distillation analyses, which include samples that have
30 atmospheric equivalent boiling points (AEBP) in the range of about
31 -44°F to 1139°F. These include ASTM Methods D2887 and D3710.
32 ASTM Method D2887 has an upper temperature limit for petroleum products
33 with a final boiling point of 1000°F at atmospheric pressure. For analysis of

1 heavier samples, such as crude oils, HTSD method D6352 extends the AEBP
2 distribution to temperatures upwards of 1300°F to 1380°F. So, the cut point of
3 the deep cut distillation of the present invention should be determined by
4 ASTM method D6352 or an equivalent method. In practice, ASTM method
5 D6352 gives distillation curve data as output and not cut point. Those skilled
6 in the art know that such distillation curve data are primary requisites for cut
7 point determination. Cut points are typically at 10 and 90 percent points on
8 distillation curves. These 10 and 90 percent points are much more reliably
9 measured than the start and end point, because they are much less
10 dependent on details of D6352 execution (such as manufacturer of equipment
11 and selection of computer software). Of course, as noted above, exact cut
12 point determination requires knowledge of adjacent cut distillation curves and
13 relative amounts.

14

15 **Hydrocracking:**

16

17 At least a portion of the light fraction from the deep cut distillation is
18 hydrocracked in a lube hydrocracking zone in the presence of a hydrocracking
19 catalyst and hydrogen under conditions to reduce the concentration of sulfur
20 and nitrogen to suitable levels for hydroisomerization dewaxing. The lube
21 hydrocracking conditions should also be selected to increase the VI to meet
22 VI specifications. Typically the solvent dewaxed VI of the feed should be
23 greater than 90 and preferably greater than 95. The concentration of nitrogen
24 in the feed for hydroisomerization dewaxing should be less than 50 ppm,
25 preferably less than 30 ppm, and more preferably less than 10 ppm. The
26 concentration of sulfur in the feed for hydroisomerization dewaxing should be
27 less than 100 ppm, preferably less than 50 ppm and more preferably less than
28 20 ppm.

29

30 Therefore, hydrocracking conditions for hydrocracking the light fraction
31 recovered from deep cut distillation, as broadly described, corresponds to
32 hydrocracking conditions for hydrocracking the waxy heavy feed stream.
33 Typical hydrocracking conditions include: reaction temperature,

1 400°F to 950°F (204°C to 510°C), preferably 650°F to 850°F
2 (343°C to 454°C); reaction pressure 500 psig to 5000 psig
3 (3.5 MPa to 34.5 MPa), preferably 1500 psig to 3500 psig
4 (10.4 MPa to 24.2 MPa); liquid hourly space velocity (LHSV),
5 0.1 hr^{-1} to 15 hr^{-1} (v/v), preferably 0.25 hr^{-1} to 2.5 hr^{-1} ; and hydrogen
6 consumption 500 to 2500 standard cubic feet per barrel of liquid hydrocarbon
7 feed ($89.1\text{--}445 \text{ m}^3 \text{ H}_2/\text{m}^3 \text{ feed}$). The hydrocracking catalyst generally
8 comprises a cracking component, a hydrogenation component, and a binder.
9 Such catalysts are well known in the art. The cracking component may
10 include an amorphous silica-alumina phase and/or a zeolite, such as a Y-type
11 or USY zeolite. The binder is generally silica, alumina or silica-alumina. The
12 hydrogenation component will be a Group VI or Group VIII metal or oxides or
13 sulfides thereof, preferably one or more of molybdenum, tungsten, cobalt, or
14 nickel, or the sulfides or oxides thereof. If present in the catalyst, these
15 hydrogenation components generally make up from about 5% to about 40%
16 by weight of the catalyst. Alternatively, platinum group metals, especially
17 platinum and/or palladium, may be present as the hydrogenation component,
18 either alone or in combination with the base metal hydrogenation components
19 molybdenum, tungsten, cobalt, or nickel. If present, the platinum group metals
20 will generally make up from about 0.1% to about 2% by weight of the catalyst.

21

22 **Hydroisomerization:**

23

24 At least a portion of the effluent from the lube hydrocracking is subjected to
25 hydroisomerization dewaxing to produce a low haze bright stock.

26

27 Hydroisomerization dewaxing catalysts useful in the present invention
28 generally comprise one or more shape selective intermediate pore size
29 molecular sieves and optionally a catalytically active metal hydrogenation
30 component on a refractory oxide support. The shape selective intermediate
31 pore size molecular sieves used alone or in combination in the practice of the
32 present invention are generally 1-D 10-, 11-, or 12-ring molecular sieves. The
33 preferred molecular sieves of the invention are of the 1-D 10-ring variety,

1 where 10-(or 11-or 12-) ring molecular sieves have 10 (or 11 or 12)
2 tetrahedrally-coordinated atoms (T-atoms) joined by oxygens. In the
3 1-D molecular sieve, the 10-ring (or larger) pores are parallel with each other,
4 and do not interconnect. The classification of intrazeolite channels as
5 1-D, 2-D and 3-D is set forth by R.M. Barrer in Zeolites,
6 Science and Technology, edited by F.R. Rodrigues, L.D. Rollman and
7 C. Naccache, NATO ASI Series, 1984 which classification is incorporated in
8 its entirety by reference (see particularly page 75).

9
10 Preferred shape selective intermediate pore size molecular sieves used for
11 hydroisomerization dewaxing are based upon aluminum phosphates, such as
12 SAPO-11, SAPO-31, and SAPO-41. SAPO-11 and SAPO-31 are more
13 preferred, with SAPO-11 being most preferred. SM-3 is a particularly
14 preferred shape selective intermediate pore size SAPO, which has a
15 crystalline structure falling within that of the SAPO-11 molecular sieves. The
16 preparation of SM-3 and its unique characteristics are described in
17 U.S. Patent Nos. 4,943,424 and 5,158,665. Also preferred shape selective
18 intermediate pore size molecular sieves used for hydroisomerization
19 dewaxing are zeolites, such as ZSM-22, ZSM-23, ZSM-35, ZSM-48, ZSM-57,
20 SSZ-32, offretite, and ferrierite. SSZ-32 and ZSM-23 are more preferred.

21
22 A particularly preferred intermediate pore size molecular sieve, which is useful
23 in the present process, is described, for example, in
24 U.S. Patent Nos. 5,135,638 and 5,282,958, the contents of which are
25 hereby incorporated by reference in their entirety. In
26 U.S. Patent No. 5,282,958, such an intermediate pore size molecular
27 sieve has a crystallite size of no more than about 0.5 microns and pores
28 with a minimum diameter of at least about 4.8 Å and with a maximum
29 diameter of about 7.1 Å. The catalyst has sufficient acidity so that 0.5 grams
30 thereof when positioned in a tube reactor converts at least 50% of
31 hexadecane at 370°C, a pressure of 1200 psig, a hydrogen flow of
32 160 ml/min, and a feed rate of 1 ml/hr. The catalyst also exhibits isomerization
33 selectivity of 40 or greater (isomerization selectivity is determined as follows:

1 100 x (weight % branched C₁₆ in product) / (weight % branched C₁₆ in product
 2 + weight % C₁₃ in product) when used under conditions leading to
 3 96% conversion of normal hexadecane (n-C₁₆) to other species.
 4
 5 Such a particularly preferred molecular sieve may further be characterized by
 6 pores or channels having a crystallographic free diameter in the range of from
 7 about 4.0 Å to about 7.1 Å, and preferably in the range of 4.0 Å to 6.5 Å. The
 8 crystallographic free diameters of the channels of molecular sieves are
 9 published in the "Atlas of Zeolite Framework Types", Fifth Revised Edition,
 10 2001, by Ch. Baerlocher, W.M. Meier, and D.H. Olson, Elsevier, pages 10-15,
 11 which is incorporated herein by reference.
 12
 13 If the crystallographic free diameters of the channels of a molecular sieve are
 14 unknown, the effective pore size of the molecular sieve can be measured
 15 using standard adsorption techniques and hydrocarbonaceous compounds of
 16 known minimum kinetic diameters. See Breck, Zeolite Molecular Sieves, 1974
 17 (especially Chapter 8); Anderson et al., J. Catalysis 58, 114 (1979); and
 18 U.S. Patent No. 4,440,871, which is incorporated herein by reference. In
 19 performing adsorption measurements to determine pore size, standard
 20 techniques are used. It is convenient to consider a particular molecule as
 21 excluded if does not reach at least 95% of its equilibrium adsorption value on
 22 the molecular sieve in less than about 10 minutes (p/po=0.5; 25°C).
 23 Intermediate pore size molecular sieves will typically admit molecules having
 24 kinetic diameters of 5.3Å to 6.5 Å with little hindrance.
 25
 26 Hydroisomerization dewaxing catalysts useful in the present invention
 27 optionally comprise a catalytically active hydrogenation metal. Typical
 28 catalytically active hydrogenation metals used alone or in combination include
 29 chromium, molybdenum, nickel, vanadium, cobalt, tungsten, zinc, platinum,
 30 and palladium. The metals platinum and palladium are especially preferred,
 31 with platinum most especially preferred. If platinum and/or palladium is used,
 32 the total amount of active hydrogenation metal is typically in the range of

1 0.1 to 5 weight % of the total catalyst, usually from 0.1 to 2 weight %, and not
2 to exceed 10 weight %.

3

4 The refractory oxide support may be selected from those oxide supports
5 which are conventionally used for catalysts, including silica, alumina,
6 silica-alumina, magnesia, titania and combinations thereof.

7

8 The catalytic hydroisomerization conditions employed depend on the feed
9 used for the hydroisomerization and the desired pour point of the product.

10 Generally, the temperature is from about 200°C to about 475°C, preferably
11 from about 250°C to about 450°C. The pressure is typically from about
12 15 psig to about 3000 psig, preferably from about 50 psig to about 2500 psig,
13 more preferably from about 100 psig to about 1000 psig, and most preferably
14 from about 150 psig to about 600 psig. The liquid hourly space
15 velocity (LHSV) is preferably from about 0.1 hr⁻¹ to about 20 hr⁻¹, more
16 preferably from about 0.1 hr⁻¹ to about 5 hr⁻¹, and most preferably from about
17 0.1 hr⁻¹ to about 1.0 hr⁻¹. Low pressure and low liquid hourly space velocity
18 provide enhanced isomerization selectivity, which results in more
19 isomerization and less cracking of the feed, thus producing an increased
20 yield.

21

22 Hydrogen is preferably present in the reaction zone during the catalytic
23 isomerization process. The hydrogen to feed ratio is typically from about
24 500 SCF/bbl to about 30,000 SCF/bbl (standard cubic feet per barrel),
25 preferably from about 1000 SCF/bbl to about 20,000 SCF/bbl.

26

27 **Hydrofinishing:**

28

29 The product from the hydroisomerization step may optionally be hydrofinished
30 in order to stabilize the lubricant product by reducing olefins and aromatics.
31 Hydrofinishing is typically conducted at temperatures ranging from about
32 300°F to about 600°F, at pressures from about 400 psig to about 3000 psig, at
33 space velocities (LHSV) from about 0.1 to about 20, and hydrogen recycle

1 rates of from about 400 SCF/bbl to about 1,500 SCF/bbl. The hydrogenation
2 catalyst employed must be active enough not only to hydrogenate the olefins,
3 diolefins and color bodies within the lube oil fractions, but also to reduce the
4 aromatic content (color bodies). The hydrofinishing step is beneficial in
5 preparing an acceptably stable lubricating oil. Suitable hydrogenation
6 catalysts include conventional metallic hydrogenation catalysts, particularly
7 the Group VIII metals such as cobalt, nickel, palladium and platinum. The
8 metals are typically associated with carriers such as bauxite, alumina, silica
9 gel, silica-alumina composites, and crystalline aluminosilicate zeolites.
10 Palladium is a particularly preferred hydrogenation metal. If desired,
11 non-noble Group VIII metals can be used with molybdates. Metal oxides or
12 sulfides can be used. Suitable catalysts are disclosed in
13 U.S. Patent Nos. 3,852,207; 4,157,294; 3,904,513 and 4,673,487, which are
14 incorporated by reference herein.

15
16 Additionally, U.S. Patent No. 6,337,010, which is incorporated by reference
17 herein discloses a process scheme for producing lubricating base oil with low
18 pressure dewaxing and high pressure hydrofinishing and discloses operating
19 conditions for lube hydrocracking, isomerization and hydrofinishing that may
20 be useful in practicing the present invention.

21

22 **Clay Treating:**

23

24 The low haze heavy base oil may optionally also be treated by a clay treating
25 step, either following hydroisomerization or following hydrofinishing to remove
26 any remaining traces of haze or haze precursors. A suitable clay treating
27 process is described, for example, in U.S. Patent No. 6,468,418, which is
28 incorporated by reference.

29

30 The following example is intended to illustrate the present invention and is not
31 intended to limit the invention in any way.

EXAMPLE

The following example describes a method of the invention for preparing low haze bright stock. An Alaska North Slope/Arabian Light/ Arabian Medium crude blend was fractionated in an atmospheric/vacuum distillation and the vacuum column bottoms upgraded by solvent deasphalting and the DAO hydrocracked. The de-asphalting process and the hydrocracking process were conventional.

The residuum (i.e., bottoms) fraction from an atmospheric fractionation of the hydrocracked DAO was topped by vacuum distillation at 700°F, and the residuum fraction separated by wiped film evaporator distillation at a 1200°F cut point. The 1200°F+ bottoms portion was 13.3 weight % of the 700°F+ fraction.

The 700°F to 1200°F distillate had the following properties:

API Gravity	20.7	
Nitrogen, ppm	988	
Sulfur, ppm	2227	
Viscosity, cSt, 100°C	21.55	
Viscosity Index	70	
Sim. Dist., weight %, °F, D6352		
	10%	767
	50%	977
	90%	1158

1 This feed was hydrocracked using a conventional commercial
 2 Ni-W-SiO₂-Al₂O₃ hydrocracking catalyst at 720/732°F reaction
 3 temperature (2 catalyst zones) at 2052 psig pressure (1968 psia H₂ pressure),
 4 0.25 hr⁻¹ feed rate, and 5200 SCFB recycle H₂ rate, and the hydrocracker
 5 effluent separated first in conventional flash separation zones and then by
 6 atmospheric/vacuum distillation. The bright stock (i.e., bottoms) fraction from
 7 the vacuum distillation was at a yield of 22.7 vol% and had the following
 8 properties:

10	API Gravity	27.1	
11	Nitrogen, ppm	7	
12	Sulfur, ppm	14	
13	Viscosity, cSt, 100°C	25.67	
14	Viscosity Index	97	
15	Wax, weight %	16.3	
16	Sim. Dist., weight %, °F, D6352		
17		10%	943
18		30%	1026
19		50%	1070
20		70%	1117
21		90%	1182
22			

23 The bright stock was converted by isomerization dewaxing at 610°F reaction
 24 temperature at 1950 psig pressure (1878 psia H₂ pressure), 1.3 hr⁻¹ feed rate,
 25 and 3000 SCFB once-through H₂ rate, over a Pt/SSZ-32 catalyst (containing
 26 35% alumina binder), followed by hydrofinishing at 450°F, 1950 psig pressure,
 27 1.0 hr⁻¹ feed rate, and 3000 SCFB once-through H₂ rate, over a
 28 Pt-Pd/SiO₂-Al₂O₃ catalyst, into a low haze bright stock lubricant product
 29 having the following properties:

31	API Gravity	27.0	
32	Viscosity, cSt		
33		(measured at 40°C)	355.7
34		(measured at 100°C)	25.68
35	Viscosity Index	95	
36	Pour Point, °C	-20°	
37	Cloud Point, °C	+2°	
38	Oxidator BN, hrs	21.5	
39	Distillation by D6352, weight %, °F		
40		10%	956
41		30%	1009
42		50%	1054
43		70%	1090
44		90%	1135

1 Lube yield was 94.7 weight %. Conducting the same set of reaction steps,
2 using a full bottoms fraction from the hydrocracked DAO would be expected to
3 produce an isomerization dewaxing/hydrofinishing product having a cloud
4 point of at least 15°C.
5
6 By contrast, solvent dewaxing the same bright stock feed produced an oil with
7 a pour point of -21°C but a VI of only 92. Furthermore, the yield of oil was only
8 83.7 weight %.